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TO ALL WHOM IT MAY CONCERN:

Be it known that WE, JOHN D. WILKINSON and HANK M. HUDSON, both citizens of the United States, both residing in Midland, County of Midland, State of Texas, whose post office addresses are 2800 W. Dengar, Midland, Texas 79705 and 2508 W. Sinclair, Midland, Texas 79705, respectively, have invented an improvement in

LIQUEFIED NATURAL GAS PROCESSING

of which the following is a

SPECIFICATION

BACKGROUND OF THE INVENTION

[0001] This invention relates to a process for the separation of ethane and heavier hydrocarbons or propane and heavier hydrocarbons from liquefied natural gas, hereinafter referred to as LNG, to provide a volatile methane-rich residue gas stream and a less volatile natural gas liquids (NGL) or liquefied petroleum gas (LPG) stream.

[0002] As an alternative to transportation in pipelines, natural gas at remote locations is sometimes liquefied and transported in special LNG tankers to appropriate LNG receiving and storage terminals. The LNG can then be re-vaporized and used as a

gaseous fuel in the same fashion as natural gas. Although LNG usually has a major proportion of methane, i.e., methane comprises at least 50 mole percent of the LNG, it also contains relatively lesser amounts of heavier hydrocarbons such as ethane, propane, butanes, and the like, as well as nitrogen. It is often necessary to separate some or all of the heavier hydrocarbons from the methane in the LNG so that the gaseous fuel resulting from vaporizing the LNG conforms to pipeline specifications for heating value. In addition, it is often also desirable to separate the heavier hydrocarbons from the methane because these hydrocarbons have a higher value as liquid products (for use as petrochemical feedstocks, as an example) than their value as fuel.

[0003] Although there are many processes which may be used to separate ethane and heavier hydrocarbons from LNG, these processes often must compromise between high recovery, low utility costs, and process simplicity (and hence low capital investment). In U.S. Patent No. 2,952,984 Marshall describes an LNG process capable of very high ethane recovery via the use of a refluxed distillation column. Markbreiter describes in U.S. Patent No. 3,837,172 a simpler process using a non-refluxed fractionation column, limited to lower ethane or propane recoveries. Rambo et al describe in U.S. Patent No. 5,114,451 an LNG process capable of very high ethane or very high propane recovery using a compressor to provide reflux for the distillation column.

[0004] The present invention is generally concerned with the recovery of ethylene, ethane, propylene, propane, and heavier hydrocarbons from such LNG streams. It uses a novel process arrangement to allow high ethane or high propane recovery while

keeping the processing equipment simple and the capital investment low. Further, the present invention offers a reduction in the utilities (power and heat) required to process the LNG to give lower operating cost than the prior art processes. A typical analysis of an LNG stream to be processed in accordance with this invention would be, in approximate mole percent, 86.7% methane, 8.9% ethane and other C_2 components, 2.9% propane and other C_3 components, and 1.0% butanes plus, with the balance made up of nitrogen.

[0005] For a better understanding of the present invention, reference is made to the following examples and drawings. Referring to the drawings:

[0006] FIGS. 1, 2, and 3 are flow diagrams of prior art LNG processing plants in accordance with United States Patent No. 3,837,172;

[0007] FIGS. 4, 5, and 6 are flow diagrams of prior art LNG processing plants in accordance with United States Patent No. 2,952,984;

[0008] FIGS. 7, 8, and 9 are flow diagrams of prior art LNG processing plants in accordance with United States Patent No. 5,114,451;

[0009] FIG. 10 is a flow diagram of an LNG processing plant in accordance with the present invention;

[0010] FIGS. 11 through 18 are flow diagrams illustrating alternative means of application of the present invention to an LNG processing plant; and

[0011] FIGS. 19 and 20 are diagrams of alternative fractionation systems which may be employed in the process of the present invention.

[0012] In the following explanation of the above figures, tables are provided summarizing flow rates calculated for representative process conditions. In the tables appearing herein, the values for flow rates (in moles per hour) have been rounded to the nearest whole number for convenience. The total stream rates shown in the tables include all non-hydrocarbon components and hence are generally larger than the sum of the stream flow rates for the hydrocarbon components. Temperatures indicated are approximate values rounded to the nearest degree. It should also be noted that the process design calculations performed for the purpose of comparing the processes depicted in the figures are based on the assumption of no heat leak from (or to) the surroundings to (or from) the process. The quality of commercially available insulating materials makes this a very reasonable assumption and one that is typically made by those skilled in the art.

[0013] For convenience, process parameters are reported in both the traditional British units and in the units of the International System of Units (SI). The molar flow rates given in the tables may be interpreted as either pound moles per hour or kilogram moles per hour. The energy consumptions reported as horsepower (HP) and/or thousand British Thermal Units per hour (MBTU/Hr) correspond to the stated molar flow rates in pound moles per hour. The energy consumptions reported as kilowatts (kW) correspond to the stated molar flow rates in kilogram moles per hour.

DESCRIPTION OF THE PRIOR ART

[0014] Referring now to FIG. 1, for comparison purposes we begin with an example of an LNG processing plant in accordance with U.S. Pat. No. 3,837,172, adapted

to produce an NGL product containing the majority of the C_2 components and heavier hydrocarbon components present in the feed stream. The LNG to be processed (stream **41**) from LNG tank **10** enters pump **11** at -255°F [-159°C]. Pump **11** elevates the pressure of the LNG sufficiently so that it can flow through heat exchangers and thence to fractionation tower **16**. Stream **41a** exiting the pump is split into two portions, streams **42** and **43**. The first portion, stream **42**, is expanded to the operating pressure (approximately 395 psia [2,723 kPa(a)]) of fractionation tower **16** by valve **12** and supplied to the tower as the top column feed.

[0015] The second portion, stream **43**, is heated prior to entering fractionation tower **16** so that all or a portion of it is vaporized, reducing the amount of liquid flowing down fractionation tower **16** and allowing the use of a smaller diameter column. In the example shown in FIG. 1, stream **43** is first heated to -229°F [-145°C] in heat exchanger **13** by cooling the liquid product from the column (stream **47**). The partially heated stream **43a** is then further heated to 30°F [-1°C] (stream **43b**) in heat exchanger **14** using a low level source of utility heat, such as the sea water used in this example. After expansion to the operating pressure of fractionation tower **16** by valve **15**, the resulting stream **43c** flows to a mid-column feed point at 27°F [-3°C].

[0016] Fractionation tower **16**, commonly referred to as a demethanizer, is a conventional distillation column containing a plurality of vertically spaced trays, one or more packed beds, or some combination of trays and packing. The trays and/or packing provide the necessary contact between the liquids falling downward in the column and the vapors rising upward. As shown in FIG. 1, the fractionation tower may consist of two

sections. The upper absorbing (rectification) section **16a** contains the trays and/or packing to provide the necessary contact between the vapors rising upward and cold liquid falling downward to condense and absorb the ethane and heavier components; the lower stripping (demethanizing) section **16b** contains the trays and/or packing to provide the necessary contact between the liquids falling downward and the vapors rising upward. The demethanizing section also includes one or more reboilers (such as reboiler **22**) which heat and vaporize a portion of the liquids flowing down the column to provide the stripping vapors which flow up the column. These vapors strip the methane from the liquids, so that the bottom liquid product (stream **47**) is substantially devoid of methane and comprised of the majority of the C_2 components and heavier hydrocarbons contained in the LNG feed stream. (Because of the temperature level required in the column reboiler, a high level source of utility heat is typically required to provide the heat input to the reboiler, such as the heating medium used in this example.) The liquid product stream **47** exits the bottom of the tower at 71°F [22°C], based on a typical specification of a methane to ethane ratio of 0.005:1 on a volume basis in the bottom product. After cooling to 19°F [-7°C] in heat exchanger **13** as described previously, the liquid product (stream **47a**) flows to storage or further processing.

[0017] The demethanizer overhead vapor, stream **46**, is the methane-rich residue gas, leaving the column at -141°F [-96°C]. After being heated to -40°F [-40°C] in cross exchanger **29** so that conventional metallurgy may be used in compressor **28**, stream **46a** enters compressor **28** (driven by a supplemental power source) and is compressed to sales line pressure (stream **46b**). Following cooling to 50°F [10°C] in cross exchanger **29**, the

residue gas product (stream **46c**) flows to the sales gas pipeline at 1315 psia [9,067 kPa(a)] for subsequent distribution.

[0018] The relative split of the LNG into streams **42** and **43** is typically adjusted to maintain the desired recovery level of the desired C_2 components and heavier hydrocarbon components in the bottom liquid product (stream **47**). Increasing the split to stream **42** feeding the top of fractionation tower **16** will increase the recovery level, until a point is reached where the composition of demethanizer overhead vapor (stream **46**) is in equilibrium with the composition of the LNG (i.e., the composition of the liquid in stream **42a**). Once this point has been reached, further increasing the split to stream **42** will not raise the recovery any further, but will simply increase the amount of high level utility heat required in reboiler **22** because less of the LNG is split to stream **43** and heated with low level utility heat in heat exchanger **14**. (High level utility heat is normally more expensive than low level utility heat, so lower operating cost is usually achieved when the use of low level heat is maximized and the use of high level heat is minimized.) For the process conditions shown in FIG. 1, the amount of LNG split to stream **42** has been set at just slightly less than this maximum amount, so that the prior art process can achieve its maximum recovery without unduly increasing the heat load in reboiler **22**.

[0019] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 1 is set forth in the following table:

Table I

(FIG. 1)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
42	4,286	440	145	49	4,941
43	5,238	537	177	60	6,038
46	9,513	54	4	0	9,618
47	11	923	318	109	1,361

Recoveries*

Ethane	94.43%
Propane	99.03%
Butanes+	99.78%

Power

LNG Feed Pump	276	HP	[454	kW]
Residue Gas Compressor	5,267	HP	[8,659	kW]
Totals	5,543	HP	[9,113	kW]

Low Level Utility Heat

LNG Heater	34,900	MBTU/Hr	[22,546	kW]

High Level Utility Heat

Demethanizer Reboiler	8,280	MBTU/Hr	[5,349	kW]

* (Based on un-rounded flow rates)

[0020] This prior art process can also be adapted to produce an LPG product containing the majority of the C_3 components and heavier hydrocarbon components present in the feed stream as shown in FIG. 2. The processing scheme for the FIG. 2 process is essentially the same as that used for the FIG. 1 process described previously. The only significant differences are that the heat input of reboiler **22** has been increased to strip the C_2 components from the liquid product (stream **47**) and the operating pressure of fractionation tower **16** has been raised slightly.

[0021] The liquid product stream **47** exits the bottom of fractionation tower **16** (commonly referred to as a deethanizer when producing an LPG product) at 189°F [87°C], based on a typical specification of an ethane to propane ratio of 0.020:1 on a molar basis in the bottom product. After cooling to 125°F [52°C] in heat exchanger **13**, the liquid product (stream **47a**) flows to storage or further processing.

[0022] The deethanizer overhead vapor (stream **46**) leaves the column at -90°F [-68°C], is heated to -40°F [-40°C] in cross exchanger **29** (stream **46a**), and is compressed by compressor **28** to sales line pressure (stream **46b**). Following cooling to 83°F [28°C] in cross exchanger **29**, the residue gas product (stream **46c**) flows to the sales gas pipeline at 1315 psia [9,067 kPa(a)] for subsequent distribution.

[0023] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 2 is set forth in the following table:

Table II

(FIG. 2)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
42	4,286	440	145	49	4,941
43	5,238	537	177	60	6,038
46	9,524	971	14	1	10,557
47	0	6	308	108	422

Recoveries*			
Propane	95.78%		
Butanes+	99.09%		
Power			
LNG Feed Pump	298 HP	[490	kW]
Residue Gas Compressor	5,107 HP	[8,396	kW]
Totals	5,405 HP	[8,886	kW]
Low Level Utility Heat			
LNG Heater	35,536 MBTU/Hr	[22,956	kW]
High Level Utility Heat			
Deethanizer Reboiler	16,525 MBTU/Hr	[10,675	kW]

* (Based on un-rounded flow rates)

[0024] If a slightly lower recovery level is acceptable, this prior art process can produce an LPG product using less power and high level utility heat as shown in FIG. 3. The processing scheme for the FIG. 3 process is essentially the same as that used for the FIG. 2 process described previously. The only significant difference is that the relative split between stream **42** and **43** has been adjusted to minimize the duty of reboiler **22** while providing the desired recovery of the C_3 components and heavier hydrocarbon components.

[0025] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 3 is set forth in the following table:

Table III

(FIG. 3)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
42	3,604	370	122	41	4,155
43	5,920	607	200	68	6,824
46	9,524	971	16	1	10,559
47	0	6	306	108	420

Recoveries*			
Propane	95.00%		
Butanes+	99.04%		
Power			
LNG Feed Pump	302	HP	[496 kW]
Residue Gas Compressor	5,034	HP	[8,276 kW]
Totals	5,336	HP	[8,772 kW]
Low Level Utility Heat			
LNG Heater	40,247	MBTU/Hr	[26,000 kW]
High Level Utility Heat			
Deethanizer Reboiler	11,827	MBTU/Hr	[7,640 kW]

* (Based on un-rounded flow rates)

[0026] FIG. 4 shows an alternative prior art process in accordance with U.S. Pat. No. 2,952,984 that can achieve higher recovery levels than the prior art process used in FIG. 1. The process of FIG. 4, adapted here to produce an NGL product containing the majority of the C_2 components and heavier hydrocarbon components present in the feed stream, has been applied to the same LNG composition and conditions as described previously for FIG. 1.

[0027] In the simulation of the FIG. 4 process, the LNG to be processed (stream **41**) from LNG tank **10** enters pump **11** at -255°F [-159°C]. Pump **11** elevates the pressure of the LNG sufficiently so that it can flow through heat exchangers and thence to fractionation tower **16**. Stream **41a** exiting the pump is heated first to -213°F [-136°C]

in reflux condenser **17** as it provides cooling to the overhead vapor (stream **46**) from fractionation tower **16**. The partially heated stream **41b** is then heated to -200°F [-129°C] (stream **41c**) in heat exchanger **13** by cooling the liquid product from the column (stream **47**), and then further heated to -137°F [-94°C] (stream **41d**) in heat exchanger **14** using low level utility heat. After expansion to the operating pressure (approximately 400 psia [2,758 kPa(a)]) of fractionation tower **16** by valve **15**, stream **41e** flows to a mid-column feed point at its bubble point, approximately -137°F [-94°C].

[0028] Overhead stream **46** leaves the upper section of fractionation tower **16** at -146°F [-99°C] and flows to reflux condenser **17** where it is cooled to -147°F [-99°C] and partially condensed by heat exchange with the cold LNG (stream **41a**) as described previously. The partially condensed stream **46a** enters reflux separator **18** wherein the condensed liquid (stream **49**) is separated from the uncondensed vapor (stream **48**). The liquid stream **49** from reflux separator **18** is pumped by reflux pump **19** to a pressure slightly above the operating pressure of demethanizer **16** and stream **49a** is then supplied as cold top column feed (reflux) to demethanizer **16**. This cold liquid reflux absorbs and condenses the C_2 components and heavier hydrocarbon components from the vapors rising in the upper rectification section of demethanizer **16**.

[0029] The liquid product stream **47** exits the bottom of fractionation tower **16** at 71°F [22°C], based on a methane to ethane ratio of 0.005:1 on a volume basis in the bottom product. After cooling to 18°F [-8°C] in heat exchanger **13** as described previously, the liquid product (stream **47a**) flows to storage or further processing. The residue gas (stream **48**) leaves reflux separator **18** at -147°F [-99°C], is heated to -40°F

[-40°C] in cross exchanger **29** (stream **48a**), and is compressed by compressor **28** to sales line pressure (stream **48b**). Following cooling to 43°F [6°C] in cross exchanger **29**, the residue gas product (stream **48c**) flows to the sales gas pipeline at 1315 psia [9,067 kPa(a)] for subsequent distribution.

[0030] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 4 is set forth in the following table:

Table IV

(FIG. 4)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
46	12,476	3	0	0	12,531
49	2,963	2	0	0	2,970
48	9,513	1	0	0	9,561
47	11	976	322	109	1,418

Recoveries*			
Ethane	99.90%		
Propane	100.00%		
Butanes+	100.00%		
Power			
LNG Feed Pump	287	HP	[472 kW]
Reflux Pump	9	HP	[15 kW]
Residue Gas Compressor	5,248	HP	[8,627 kW]
Totals	5,544	HP	[9,114 kW]
Low Level Utility Heat			
LNG Heater	11,265	MBTU/Hr	[7,277 kW]
High Level Utility Heat			
Demethanizer Reboiler	30,968	MBTU/Hr	[20,005 kW]

* (Based on un-rounded flow rates)

[0031] Comparing the recovery levels displayed in Table IV above for the FIG. 4 prior art process with those in Table I for the FIG. 1 prior art process shows that the FIG. 4 process can achieve substantially higher ethane, propane, and butanes+ recoveries. However, comparing the utilities consumptions in Table IV with those in Table I shows that the high level utility heat required for the FIG. 4 process is much higher than that for the FIG. 1 process because the FIG. 4 process does not allow for optimum use of low level utility heat.

[0032] This prior art process can also be adapted to produce an LPG product containing the majority of the C_3 components and heavier hydrocarbon components present in the feed stream as shown in FIG. 5. The processing scheme for the FIG. 5 process is essentially the same as that used for the FIG. 4 process described previously. The only significant differences are that the heat input of reboiler **22** has been increased to strip the C_2 components from the liquid product (stream **47**) and the operating pressure of fractionation tower **16** has been raised slightly. The LNG composition and conditions are the same as described previously for FIG. 2.

[0033] The liquid product stream **47** exits the bottom of deethanizer **16** at 190°F [88°C], based on an ethane to propane ratio of 0.020:1 on a molar basis in the bottom product. After cooling to 125°F [52°C] in heat exchanger **13**, the liquid product (stream **47a**) flows to storage or further processing. The residue gas (stream **48**) leaves reflux separator **18** at -94°F [-70°C], is heated to -40°F [-40°C] in cross exchanger **29** (stream **48a**), and is compressed by compressor **28** to sales line pressure (stream **48b**). Following cooling to 79°F [26°C] in cross exchanger **29**, the residue gas product (stream **48c**) flows to the sales gas pipeline at 1315 psia [9,067 kPa(a)] for subsequent distribution.

[0034] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 5 is set forth in the following table:

Table V

(FIG. 5)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
46	11,401	2,783	3	0	14,238
49	1,877	1,812	3	0	3,696
48	9,524	971	0	0	10,542
47	0	6	322	109	437

Recoveries*			
Propane	99.90%		
Butanes+	100.00%		
Power			
LNG Feed Pump	309 HP	[508	kW]
Reflux Pump	12 HP	[20	kW]
Residue Gas Compressor	5,106 HP	[8,394	kW]
Totals	5,427 HP	[8,922	kW]
Low Level Utility Heat			
LNG Heater	1,689 MBTU/Hr	[1,091	kW]
High Level Utility Heat			
Deethanizer Reboiler	49,883 MBTU/Hr	[32,225	kW]

* (Based on un-rounded flow rates)

[0035] If a slightly lower recovery level is acceptable, this prior art process can produce an LPG product using less power and high level utility heat as shown in FIG. 6. The processing scheme for the FIG. 6 process is essentially the same as that used for the FIG. 5 process described previously. The only significant difference is that the outlet temperature of stream **46a** from reflux condenser **17** has been adjusted to minimize the duty of reboiler **22** while providing the desired recovery of the C_3 components and heavier hydrocarbon components. The LNG composition and conditions are the same as described previously for FIG. 3.

[0036] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 6 is set forth in the following table:

Table VI

(FIG. 6)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
46	10,485	1,910	97	0	12,541
49	961	939	81	0	1,983
48	9,524	971	16	0	10,558
47	0	6	306	109	421

Recoveries*			
Propane	95.00%		
Butanes+	100.00%		
Power			
LNG Feed Pump	309	HP	[508 kW]
Reflux Pump	7	HP	[12 kW]
Residue Gas Compressor	5,108	HP	[8,397 kW]
Totals	5,424	HP	[8,917 kW]
Low Level Utility Heat			
LNG Heater	8,230	MBTU/Hr	[5,317 kW]
High Level Utility Heat			
Deethanizer Reboiler	43,768	MBTU/Hr	[28,274 kW]

* (Based on un-rounded flow rates)

[0037] FIG. 7 shows another alternative prior art process in accordance with U.S. Pat. No. 5,114,451 that can also achieve higher recovery levels than the prior art process used in FIG. 1. The process of FIG. 7, adapted here to produce an NGL product containing the majority of the C_2 components and heavier hydrocarbon components present in the feed stream, has been applied to the same LNG composition and conditions as described previously for FIGS. 1 and 4.

[0038] In the simulation of the FIG. 7 process, the LNG to be processed (stream **41**) from LNG tank **10** enters pump **11** at -255°F [-159°C]. Pump **11** elevates the pressure of the LNG sufficiently so that it can flow through heat exchangers and thence

to fractionation tower **16**. Stream **41a** exiting the pump is split into two portions, streams **42** and **43**. The second portion, stream **43**, is heated prior to entering fractionation tower **16** so that all or a portion of it is vaporized, reducing the amount of liquid flowing down fractionation tower **16** and allowing the use of a smaller diameter column. In the example shown in FIG. 7, stream **43** is first heated to -226°F [-143°C] in heat exchanger **13** by cooling the liquid product from the column (stream **47**). The partially heated stream **43a** is then further heated to 30°F [-1°C] (stream **43b**) in heat exchanger **14** using low level utility heat. After expansion to the operating pressure (approximately 395 psia [2,723 kPa(a)]) of fractionation tower **16** by valve **15**, stream **43c** flows to a lower mid-column feed point at 27°F [-3°C].

[0039] The proportion of the total feed in stream **41a** flowing to the column as stream **42** is controlled by valve **12**, and is typically 50% or less of the total feed. Stream **42a** flows from valve **12** to heat exchanger **17** where it is heated as it cools, substantially condenses, and subcools stream **49a**. The heated stream **42b** then flows to demethanizer **16** at an upper mid-column feed point at -160°F [-107°C].

[0040] Tower overhead stream **46** leaves demethanizer **16** at -147°F [-99°C] and is divided into two portions. The major portion, stream **48**, is the methane-rich residue gas. It is heated to -40°F [-40°C] in cross exchanger **29** (stream **48a**) and compressed by compressor **28** to sales line pressure (stream **48b**). Following cooling to 43°F [6°C] in cross exchanger **29**, the residue gas product (stream **48c**) flows to the sales gas pipeline at 1315 psia [9,067 kPa(a)] for subsequent distribution.

[0041] The minor portion of the tower overhead, stream **49**, enters compressor **26**, which supplies a modest boost in pressure to overcome the pressure drops in heat exchanger **17** and control valve **27**, as well as the static head due to the height of demethanizer **16**. The compressed stream **49a** is cooled to -247°F [-155°C] to substantially condense and subcool it (stream **49b**) by a portion of the LNG feed (stream **42a**) in heat exchanger **17** as described previously. Stream **49b** flows through valve **27** to lower its pressure to that of fractionation tower **16**, and resulting stream **49c** flows to the top feed point of demethanizer **16** to serve as reflux for the tower.

[0042] The liquid product stream **47** exits the bottom of fractionation tower **16** at 70°F [21°C], based on a methane to ethane ratio of 0.005:1 on a volume basis in the bottom product. After cooling to 18°F [-8°C] in heat exchanger **13** as described previously, the liquid product (stream **47a**) flows to storage or further processing.

[0043] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 7 is set forth in the following table:

Table VII

(FIG. 7)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
42	4,762	488	161	54	5,489
43	4,762	489	161	55	5,490
46	11,503	1	0	0	11,561
49	1,990	0	0	0	2,000
48	9,513	1	0	0	9,561
47	11	976	322	109	1,418

Recoveries*			
Ethane	99.88%		
Propane	100.00%		
Butanes+	100.00%		
Power			
LNG Feed Pump	276	HP	[454 kW]
Recycle Compressor	48	HP	[79 kW]
Residue Gas Compressor	5,249	HP	[8,629 kW]
Totals	5,573	HP	[9,162 kW]
Low Level Utility Heat			
LNG Heater	31,489	MBTU/Hr	[20,342 kW]
High Level Utility Heat			
Demethanizer Reboiler	10,654	MBTU/Hr	[6,883 kW]

* (Based on un-rounded flow rates)

[0044] Comparing the recovery levels displayed in Table VII above for the FIG. 7 prior art process with those in Table I for the FIG. 1 prior art process shows that the FIG. 7 process can achieve substantially higher ethane, propane, and butanes+ recoveries, essentially the same as those achieved by the FIG. 4 prior art process as shown in Table IV. Further, comparing the utilities consumptions in Table VII with those in Table IV shows that the high level utility heat required for the FIG. 7 process is much lower than that for the FIG. 4. In fact, the high level utility heat required for the FIG. 7 process is only about 29% higher than the FIG. 1 process.

[0045] This prior art process can also be adapted to produce an LPG product containing the majority of the C_3 components and heavier hydrocarbon components present in the feed stream as shown in FIG. 8. The processing scheme for the FIG. 8 process is essentially the same as that used for the FIG. 7 process described previously. The only significant differences are that the heat input of reboiler **22** has been increased to strip the C_2 components from the liquid product (stream **47**), the relative split between stream **42** and **43** has been adjusted to minimize the duty of reboiler **22** while providing the desired recovery of the C_3 components and heavier hydrocarbon components, and the operating pressure of fractionation tower **16** has been raised slightly. The LNG composition and conditions are the same as described previously for FIGS. 2 and 5.

[0046] The liquid product stream **47** exits the bottom of deethanizer **16** at 189°F [87°C], based on an ethane to propane ratio of 0.020:1 on a molar basis in the bottom product. After cooling to 124°F [51°C] in heat exchanger **13**, the liquid product (stream **47a**) flows to storage or further processing. The residue gas (stream **48**) at -93°F [-70°C] is heated to -40°F [-40°C] in cross exchanger **29** (stream **48a**) and compressed by compressor **28** to sales line pressure (stream **48b**). Following cooling to 78°F [25°C] in cross exchanger **29**, the residue gas product (stream **48c**) flows to the sales gas pipeline at 1315 psia [9,067 kPa(a)] for subsequent distribution.

[0047] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 8 is set forth in the following table:

Table VIII

(FIG. 8)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
42	5,714	586	193	65	6,587
43	3,810	391	129	44	4,392
46	12,676	1,292	0	0	14,032
49	3,152	321	0	0	3,490
48	9,524	971	0	0	10,542
47	0	6	322	109	437

Recoveries*					
Propane	99.90%				
Butanes+	100.00%				
Power					
LNG Feed Pump	302	HP	[	496	kW]
Recycle Compressor	104	HP	[	171	kW]
Residue Gas Compressor	5,033	HP	[	8,274	kW]
Totals	5,439	HP	[	8,941	kW]
Low Level Utility Heat					
LNG Heater	25,468	MBTU/Hr	[	16,452	kW]
High Level Utility Heat					
Demethanizer Reboiler	25,808	MBTU/Hr	[	16,672	kW]

* (Based on un-rounded flow rates)

[0048] If a slightly lower recovery level is acceptable, this prior art process can produce an LPG product using less power and high level utility heat as shown in FIG. 9. The processing scheme for the FIG. 9 process is essentially the same as that used for the FIG. 8 process described previously. The only significant differences are that the relative split between stream **42** and **43** and the flow rate of recycle stream **49** have been adjusted to minimize the duty of reboiler **22** while providing the desired recovery of the C_3 components and heavier hydrocarbon components. The LNG composition and conditions are the same as described previously for FIGS. 3 and 6.

[0049] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 9 is set forth in the following table:

Table IX

(FIG. 9)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
42	4,374	449	148	50	5,042
43	5,150	528	174	59	5,937
46	11,327	1,155	19	0	12,558
49	1,803	184	3	0	2,000
48	9,524	971	16	0	10,558
47	0	6	306	109	421

Recoveries*			
Propane	95.00%		
Butanes+	100.00%		
Power			
LNG Feed Pump	302	HP	[496 kW]
Recycle Compressor	61	HP	[100 kW]
Residue Gas Compressor	5,034	HP	[8,276 kW]
Totals	5,397	HP	[8,872 kW]
Low Level Utility Heat			
LNG Heater	34,868	MBTU/Hr	[22,525 kW]
High Level Utility Heat			
Demethanizer Reboiler	16,939	MBTU/Hr	[10,943 kW]

* (Based on un-rounded flow rates)

DESCRIPTION OF THE INVENTION

Example 1

[0050] FIG. 10 illustrates a flow diagram of a process in accordance with the present invention. The LNG composition and conditions considered in the process presented in FIG. 10 are the same as those in FIGS. 1, 4, and 7. Accordingly, the FIG. 10 process can be compared with that of the FIGS. 1, 4, and 7 processes to illustrate the advantages of the present invention.

[0051] In the simulation of the FIG. 10 process, the LNG to be processed (stream **41**) from LNG tank **10** enters pump **11** at -255°F [-159°C]. Pump **11** elevates the pressure of the LNG sufficiently so that it can flow through heat exchangers and thence to fractionation tower **16**. Stream **41a** exiting the pump is heated to -152°F [-102°C] in reflux condenser **17** as it provides cooling to the overhead vapor (stream **46**) from fractionation tower **16**. Stream **41b** exiting reflux condenser **17** is split into two portions, streams **42** and **43**. The first portion, stream **42**, is expanded to the operating pressure (approximately 400 psia [2,758 kPa(a)]) of fractionation tower **16** by valve **12** and supplied to the tower at an upper mid-column feed point.

[0052] The second portion, stream **43**, is heated prior to entering fractionation tower **16** so that all or a portion of it is vaporized, reducing the amount of liquid flowing down fractionation tower **16** and allowing the use of a smaller diameter column. In the example shown in FIG. 10, stream **43** is first heated to -137°F [-94°C] in heat exchanger **13** by cooling the liquid product from the column (stream **47**). The partially heated stream **43a** is then further heated to 30°F [-1°C] (stream **43b**) in heat exchanger **14** using low level utility heat. After expansion to the operating pressure of fractionation tower **16** by valve **15**, stream **43c** flows to a lower mid-column feed point at 27°F [-3°C].

[0053] The demethanizer in fractionation tower **16** is a conventional distillation column containing a plurality of vertically spaced trays, one or more packed beds, or some combination of trays and packing. As shown in FIG. 10, the fractionation tower may consist of two sections. The upper absorbing (rectification) section **16a** contains the trays and/or packing to provide the necessary contact between the vapors rising upward

and cold liquid falling downward to condense and absorb the ethane and heavier components; the lower stripping (demethanizing) section **16b** contains the trays and/or packing to provide the necessary contact between the liquids falling downward and the vapors rising upward. The demethanizing section also includes one or more reboilers (such as reboiler **22**) which heat and vaporize a portion of the liquids flowing down the column to provide the stripping vapors which flow up the column. The liquid product stream **47** exits the bottom of the tower at 71°F [22°C], based on a methane to ethane ratio of 0.005:1 on a volume basis in the bottom product. After cooling to 18°F [-8°C] in heat exchanger **13** as described previously, the liquid product (stream **47a**) flows to storage or further processing.

[0054] Overhead distillation stream **46** is withdrawn from the upper section of fractionation tower **16** at -146°F [-99°C] and flows to reflux condenser **17** where it is cooled to -147°F [-99°C] and partially condensed by heat exchange with the cold LNG (stream **41a**) as described previously. The partially condensed stream **46a** enters reflux separator **18** wherein the condensed liquid (stream **49**) is separated from the uncondensed vapor (stream **48**). The liquid stream **49** from reflux separator **18** is pumped by reflux pump **19** to a pressure slightly above the operating pressure of demethanizer **16** and stream **49a** is then supplied as cold top column feed (reflux) to demethanizer **16**. This cold liquid reflux absorbs and condenses the C_2 components and heavier hydrocarbon components from the vapors rising in the upper rectification section of demethanizer **16**.

[0055] The residue gas (stream **48**) leaves reflux separator **18** at -147°F [-99°C], is heated to -40°F [-40°C] in cross exchanger **29** (stream **48a**), and is compressed by

compressor **28** to sales line pressure (stream **48b**). Following cooling to 43°F [6°C] in cross exchanger **29**, the residue gas product (stream **48c**) flows to the sales gas pipeline at 1315 psia [9,067 kPa(a)] for subsequent distribution.

[0056] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 10 is set forth in the following table:

Table X

(FIG. 10)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
42	3,048	313	103	35	3,513
43	6,476	664	219	74	7,466
46	17,648	8	0	0	17,717
49	8,135	7	0	0	8,156
48	9,513	1	0	0	9,561
47	11	976	322	109	1,418

Recoveries*			
Ethane	99.90%		
Propane	100.00%		
Butanes+	100.00%		
Power			
LNG Feed Pump	287	HP	[472 kW]
Reflux Pump	25	HP	[41 kW]
Residue Gas Compressor	5,248	HP	[8,628 kW]
Totals	5,560	HP	[9,141 kW]
Low Level Utility Heat			
LNG Heater	32,493	MBTU/Hr	[20,991 kW]
High Level Utility Heat			
Demethanizer Reboiler	9,741	MBTU/Hr	[6,293 kW]

* (Based on un-rounded flow rates)

[0057] Comparing the recovery levels displayed in Table X above for the FIG. 10 process with those in Table I for the FIG. 1 prior art process shows that the present invention can achieve much higher liquids recovery efficiency than the FIG. 1 process. Comparing the utilities consumptions in Table X with those in Table I shows that the power requirement for the present invention is essentially the same as that for the FIG. 1 process, and that the high level utility heat required for the present invention is only slightly higher (about 18%) than that for the FIG. 1 process.

[0058] Comparing the recovery levels displayed in Table X with those in Tables IV and VII for the FIGS. 4 and 7 prior art processes shows that the present invention matches the liquids recovery efficiencies of the FIGS. 4 and 7 processes. Comparing the utilities consumptions in Table X with those in Tables IV and VII shows that the power requirement for the present invention is essentially the same as that for the FIGS. 4 and 7 processes, but that the high level utility heat required for the present invention is substantially lower (about 69% lower and 9% lower, respectively) than that for the FIGS. 4 and 7 processes.

[0059] There are three primary factors that account for the improved efficiency of the present invention. First, compared to the FIG. 1 prior art process, the present invention does not depend on the LNG feed itself to directly serve as the reflux for fractionation column **16**. Rather, the refrigeration inherent in the cold LNG is used indirectly in reflux condenser **17** to generate a liquid reflux stream (stream **49**) that contains very little of the C_2 components and heavier hydrocarbon components that are to be recovered, resulting in efficient rectification in the upper absorbing section **16a** of fractionation tower **16** and avoiding the equilibrium limitations of the prior art FIG. 1 process (similar to the steps shown in the FIG. 4 prior art process). Second, compared to the FIG. 4 prior art process, splitting the LNG feed into two portions before feeding fractionation tower **16** allows more efficient use of low level utility heat, thereby reducing the amount of high level utility heat consumed by reboiler **22**. The relatively colder portion of the LNG feed (stream **42a** in FIG. 10) serves as a second reflux stream for fractionation tower **16**, providing partial rectification of the vapors in the heated

portion (stream **43c** in FIG. 10) so that heating and vaporizing this portion of the LNG feed does not unduly increase the load on reflux condenser **17**. Third, compared to the FIG. 7 prior art process, using the entire LNG feed (stream **41a** in FIG. 10) in reflux condenser **17** rather than just a portion (stream **42a** in FIG. 7) allows generating more reflux for fractionation tower **16**, as can be seen by comparing stream **49** in Table X with stream **49** in Table VII. The higher reflux flow allows more of the LNG feed to be heated using low level utility heat in heat exchanger **14** (compare stream **43** in Table X with stream **43** in Table VII), reducing the duty required in reboiler **22** and minimizing the amount of high level utility heat needed to meet the specification for the bottom liquid product from the demethanizer.

Example 2

[0060] The present invention can also be adapted to produce an LPG product containing the majority of the C_3 components and heavier hydrocarbon components present in the feed stream as shown in FIG. 11. The LNG composition and conditions considered in the process presented in FIG. 11 are the same as described previously for FIGS. 2, 5, and 8. Accordingly, the FIG. 11 process of the present invention can be compared to the prior art processes displayed in FIGS. 2, 5, and 8.

[0061] The processing scheme for the FIG. 11 process is essentially the same as that used for the FIG. 10 process described previously. The only significant differences are that the heat input of reboiler **22** has been increased to strip the C_2 components from the liquid product (stream **47**) and the operating pressure of fractionation tower **16** has been raised slightly.

[0062] The liquid product stream **47** exits the bottom of deethanizer **16** at 189°F [87°C], based on an ethane to propane ratio of 0.020:1 on a molar basis in the bottom product. After cooling to 124°F [51°C] in heat exchanger **13**, the liquid product (stream **47a**) flows to storage or further processing. The residue gas (stream **48**) leaves reflux separator **18** at -94°F [-70°C], is heated to -40°F [-40°C] in cross exchanger **29** (stream **48a**), and is compressed by compressor **28** to sales line pressure (stream **48b**). Following cooling to 79°F [26°C] in cross exchanger **29**, the residue gas product (stream **48c**) flows to the sales gas pipeline at 1315 psia [9,067 kPa(a)] for subsequent distribution.

[0063] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 11 is set forth in the following table:

Table XI

(FIG. 11)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
42	3,048	313	103	35	3,513
43	6,476	664	219	74	7,466
46	12,067	3,425	4	0	15,547
49	2,543	2,454	4	0	5,005
48	9,524	971	0	0	10,542
47	0	6	322	109	437

Recoveries*					
Propane	99.90%				
Butanes+	100.00%				
Power					
LNG Feed Pump	309	HP	[	508	kW]
Reflux Pump	16	HP	[	26	kW]
Residue Gas Compressor	5,106	HP	[	8,394	kW]
Totals	5,431	HP	[	8,928	kW]
Low Level Utility Heat					
LNG Heater	28,486	MBTU/Hr	[	18,402	kW]
High Level Utility Heat					
Deethanizer Reboiler	23,077	MBTU/Hr	[	14,908	kW]

* (Based on un-rounded flow rates)

[0064] Comparing the recovery levels displayed in Table XI above for the FIG. 11 process with those in Table II for the FIG. 2 prior art process shows that the present invention can achieve much higher liquids recovery efficiency than the FIG. 2 process. Comparing the utilities consumptions in Table XI with those in Table II shows that the power requirement for the present invention is essentially the same as that for the FIG. 2 process, although the high level utility heat required for the present invention is significantly higher (about 40%) than that for the FIG. 2 process.

[0065] Comparing the recovery levels displayed in Table XI with those in Tables V and VIII for the FIGS. 5 and 8 prior art processes shows that the present

invention matches the liquids recovery efficiencies of the FIGS. 5 and 8 processes. Comparing the utilities consumptions in Table XI with those in Tables V and VIII shows that the power requirement for the present invention is essentially the same as that for the FIGS. 5 and 8 processes, but that the high level utility heat required for the present invention is substantially lower (about 54% lower and 11% lower, respectively) than that for the FIGS. 5 and 8 processes.

Example 3

[0066] If a slightly lower recovery level is acceptable, another embodiment of the present invention may be employed to produce an LPG product using much less power and high level utility heat. FIG. 12 illustrates such an alternative embodiment. The LNG composition and conditions considered in the process presented in FIG. 12 are the same as those in FIG. 11, as well as those described previously for FIGS. 3, 6, and 9. Accordingly, the FIG. 12 process of the present invention can be compared to the embodiment displayed in FIG. 11 and to the prior art processes displayed in FIGS. 3, 6, and 9.

[0067] In the simulation of the FIG. 12 process, the LNG to be processed (stream **41**) from LNG tank **10** enters pump **11** at -255°F [-159°C]. Pump **11** elevates the pressure of the LNG sufficiently so that it can flow through heat exchangers and thence to absorber column **16**. Stream **41a** exiting the pump is heated first to -91°F [-69°C] in reflux condenser **17** as it provides cooling to the overhead vapor (distillation stream **46**) withdrawn from contacting device absorber column **16**. The partially heated stream **41b** is then heated to -88°F [-67°C] (stream **41c**) in heat exchanger **13** by cooling the liquid

product (stream **47**) from fractionation stripper column **21**, and then further heated to 30°F [-1°C] (stream **41d**) in heat exchanger **14** using low level utility heat. After expansion to the operating pressure (approximately 855 psia [5,895 kPa(a)]) of absorber column **16** by valve **15**, stream **41e** flows to a lower column feed point on the column at 28°F [-2°C]. The liquid portion (if any) of expanded stream **41e** commingles with liquids falling downward from the upper section of absorber column **16** and the combined liquid stream **44** exits the bottom of contacting device absorber column **16** at 17°F [-8°C]. The vapor portion of expanded stream **41e** rises upward through absorber column **16** and is contacted with cold liquid falling downward to condense and absorb the C_3 components and heavier hydrocarbon components.

[0068] The combined liquid stream **44** from the bottom of the absorber column **16** is flash expanded to slightly above the operating pressure (430 psia [2,965 kPa(a)]) of stripper column **21** by expansion valve **20**, cooling stream **44** to -11°F [-24°C] (stream **44a**) before it enters fractionation stripper column **21** at a top column feed point. In the stripper column **21**, stream **44a** is stripped of its methane and C_2 components by the vapors generated in reboiler **22** to meet the specification of an ethane to propane ratio of 0.020:1 on a molar basis. The resulting liquid product stream **47** exits the bottom of stripper column **21** at 191°F [88°C] and is cooled to 126°F [52°C] in heat exchanger **13** (stream **47a**) before flowing to storage or further processing.

[0069] The overhead vapor (stream **45**) from stripper column **21** exits the column at 52°F [11°C] and enters overhead compressor **23** (driven by a supplemental power source). Overhead compressor **23** elevates the pressure of stream **45a** to slightly above

the operating pressure of absorber column **16** so that stream **45a** can be supplied to absorber column **16** at a lower column feed point. Stream **45a** enters absorber column **16** at 144°F [62°C], whereupon it rises upward through absorber column **16** and is contacted with cold liquid falling downward to condense and absorb the C_3 components and heavier hydrocarbon components.

[0070] Overhead distillation stream **46** is withdrawn from contacting device absorber column **16** at -63°F [-53°C] and flows to reflux condenser **17** where it is cooled to -78°F [-61°C] and partially condensed by heat exchange with the cold LNG (stream **41a**) as described previously. The partially condensed stream **46a** enters reflux separator **18** wherein the condensed liquid (stream **49**) is separated from the uncondensed vapor (stream **48**). The liquid stream **49** from reflux separator **18** is pumped by reflux pump **19** to a pressure slightly above the operating pressure of absorber column **16** and stream **49a** is then supplied as cold top column feed (reflux) to absorber column **16**. This cold liquid reflux absorbs and condenses the C_3 components and heavier hydrocarbon components from the vapors rising in absorber column **16**.

[0071] The residue gas (stream **48**) leaves reflux separator **18** at -78°F [-61°C], is heated to -40°F [-40°C] in cross exchanger **29** (stream **48a**), and is compressed by compressor **28** to sales line pressure (stream **48b**). Following cooling to -37°F [-38°C] in cross exchanger **29**, stream **48c** is heated to 30°F [-1°C] using low level utility heat in heat exchanger **30** and the residue gas product (stream **48d**) flows to the sales gas pipeline at 1315 psia [9,067 kPa(a)] for subsequent distribution.

[0072] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 12 is set forth in the following table:

Table XII

(FIG. 12)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
44	705	447	552	129	1,835
45	705	441	246	20	1,414
46	31,114	4,347	93	0	35,687
49	21,590	3,376	77	0	25,129
48	9,524	971	16	0	10,558
47	0	6	306	109	421

Recoveries*		
Propane	95.01%	
Butanes+	99.98%	
Power		
LNG Feed Pump	616 HP	[1,013 kW]
Reflux Pump	117 HP	[192 kW]
Overhead Compressor	422 HP	[694 kW]
Residue Gas Compressor	1,424 HP	[2,341 kW]
Totals	2,579 HP	[4,240 kW]
Low Level Utility Heat		
LNG Heater	32,436 MBTU/Hr	[20,954 kW]
Residue Gas Heater	12,541 MBTU/Hr	[8,101 kW]
Totals	44,977 MBTU/Hr	[29,055 kW]
High Level Utility Heat		
Deethanizer Reboiler	7,336 MBTU/Hr	[4,739 kW]

* (Based on un-rounded flow rates)

[0073] Comparing Table XII above for the FIG. 12 embodiment of the present invention with Table XI for the FIG. 11 embodiment of the present invention shows that there is a reduction in liquids recovery (from 99.90% propane recovery and 100.00% butanes+ recovery to 95.01% propane recovery and 99.98% butanes+ recovery) for the FIG. 12 embodiment. However, the power and heat requirements for the FIG. 12 embodiment are less than one-half of those for the FIG. 11 embodiment. The choice of

which embodiment to use for a particular application will generally be dictated by the monetary value of the heavier hydrocarbons in the LPG product versus their corresponding value as gaseous fuel in the residue gas product, and by the cost of power and high level utility heat.

[0074] Comparing the recovery levels displayed in Table XII with those in Tables III, VI, and IX for the FIGS. 3, 6, and 9 prior art processes shows that the present invention matches the liquids recovery efficiencies of the FIGS. 3, 6, and 9 processes. Comparing the utilities consumptions in Table XII with those in Tables III, VI, and IX shows that the power requirement for this embodiment of the present invention is significantly less (about 52% lower) than that for the FIGS. 3, 6, and 9 processes, as is the high level utility heat required (about 38%, 83%, and 57% lower, respectively, than that for the FIGS. 3, 6, and 9 processes).

[0075] Comparing this embodiment of the present invention to the prior art process displayed in FIGS. 3, 6, and 9, note that while the operating pressure of fractionation stripper column **21** is the same as that of fractionation tower **16** in the three prior art processes, the operating pressure of contacting device absorber column **16** is significantly higher, 855 psia [5,895 kPa(a)] versus 430 psia [2,965 kPa(a)]. Accordingly, the residue gas enters compressor **28** at a higher pressure in the FIG. 12 embodiment of the present invention and less compression horsepower is therefore needed to deliver the residue gas to pipeline pressure.

[0076] Since the prior art processes perform rectification and stripping in the same tower (i.e., absorbing section **16a** and stripping section **16b** contained in

fractionation tower **16** in FIG. 1), the two operations must of necessity be performed at essentially the same pressure in the prior art processes. The power consumption of the prior art processes could be reduced by raising the operating pressure of deethanizer **16**. Unfortunately, this is not advisable in this instance because of the detrimental effect on distillation performance in deethanizer **16** that would result from the higher operating pressure. This effect is manifested by poor mass transfer in deethanizer **16** due to the phase behavior of its vapor and liquid streams. Of particular concern are the physical properties that affect the vapor-liquid separation efficiency, namely the liquid surface tension and the differential in the densities of the two phases. As a result, the operating pressure of deethanizer **16** should not be raised above the values shown in FIGS. 3, 6, and 9, so there is no means available to reduce the power consumption of compressor **28** using the prior art process.

[0077] With overhead compressor **23** supplying the motive force to cause the overhead from stripper column **21** (stream **45** in FIG. 12) to flow to absorber column **16**, the operating pressures of the rectification operation (absorber column **16**) and the stripping operation (stripper column **21**) are no longer coupled together as they are in the prior art processes. Instead, the operating pressures of the two columns can be optimized independently. In the case of stripper column **21**, the pressure can be selected to insure good distillation characteristics, while for absorber column **16** the pressure can be selected to optimize the liquids recovery level versus the residue gas compression power requirements.

[0078] The dramatic reduction in the duty of reboiler **22** for the FIG. 12 embodiment of the present invention is the result of two factors. First, as liquid stream **44** from the bottom of absorber column **16** is flash expanded to the operating pressure of stripper column **21**, a significant portion of the methane and C_2 components in this stream is vaporized. These vapors return to absorber column **16** in stream **45a** to serve as stripping vapors for the liquids flowing downward in the absorber column, so that there is less of the methane and C_2 components to be stripped from the liquids in stripper column **21**. Second, overhead compressor **23** is in essence a heat pump serving as a side reboiler to absorber column **16**, since the heat of compression is supplied directly to the bottom of absorber column **16**. This further reduces the amount of methane and C_2 components contained in stream **44** that must be stripped from the liquids in stripper column **21**.

Example 4

[0079] A slightly more complex design that maintains the same C_3 component recovery with lower power consumption can be achieved using another embodiment of the present invention as illustrated in the FIG. 13 process. The LNG composition and conditions considered in the process presented in FIG. 13 are the same as those in FIG. 12. Accordingly, the FIG. 13 embodiment can be compared to the embodiment displayed in FIG. 12.

[0080] In the simulation of the FIG. 13 process, the LNG to be processed (stream **41**) from LNG tank **10** enters pump **11** at -255°F [-159°C]. Pump **11** elevates the pressure of the LNG sufficiently so that it can flow through heat exchangers and thence to absorber column **16**. Stream **41a** exiting the pump is heated first to -104°F [-76°C] in

reflux condenser **17** as it provides cooling to the overhead vapor (distillation stream **46**) withdrawn from contacting device absorber column **16**. The partially heated stream **41b** is then heated to -88°F [-67°C] (stream **41c**) in heat exchanger **13** by cooling the overhead stream (stream **45a**) and the liquid product (stream **47**) from fractionation stripper column **21**, and then further heated to 30°F [-1°C] (stream **41d**) in heat exchanger **14** using low level utility heat. After expansion to the operating pressure (approximately 855 psia [5,895 kPa(a)]) of absorber column **16** by valve **15**, stream **41e** flows to a lower column feed point on absorber column **16** at 28°F [-2°C]. The liquid portion (if any) of expanded stream **41e** commingles with liquids falling downward from the upper section of absorber column **16** and the combined liquid stream **44** exits the bottom of absorber column **16** at 5°F [-15°C]. The vapor portion of expanded stream **41e** rises upward through absorber column **16** and is contacted with cold liquid falling downward to condense and absorb the C_3 components and heavier hydrocarbon components.

[0081] The combined liquid stream **44** from the bottom of contacting device absorber column **16** is flash expanded to slightly above the operating pressure (430 psia [2,965 kPa(a)]) of stripper column **21** by expansion valve **20**, cooling stream **44** to -24°F [-31°C] (stream **44a**) before it enters fractionation stripper column **21** at a top column feed point. In the stripper column **21**, stream **44a** is stripped of its methane and C_2 components by the vapors generated in reboiler **22** to meet the specification of an ethane to propane ratio of 0.020:1 on a molar basis. The resulting liquid product stream **47** exits

the bottom of stripper column **21** at 191°F [88°C] and is cooled to 126°F [52°C] in heat exchanger **13** (stream **47a**) before flowing to storage or further processing.

[0082] The overhead vapor (stream **45**) from stripper column **21** exits the column at 43°F [6°C] and flows to cross exchanger **24** where it is cooled to -47°F [-44°C] and partially condensed. Partially condensed stream **45a** is further cooled to -99°F [-73°C] in heat exchanger **13** as previously described, condensing the remainder of the stream. Condensed liquid stream **45b** then enters overhead pump **25**, which elevates the pressure of stream **45c** to slightly above the operating pressure of absorber column **16**. Stream **45c** returns to cross exchanger **24** and is heated to 38°F [3°C] and partially vaporized as it provides cooling to stream **45**. Partially vaporized stream **45d** is then supplied to absorber column **16** at a lower column feed point, whereupon its vapor portion rises upward through absorber column **16** and is contacted with cold liquid falling downward to condense and absorb the C_3 components and heavier hydrocarbon components. The liquid portion of stream **45d** commingles with liquids falling downward from the upper section of absorber column **16** and becomes part of combined liquid stream **44** leaving the bottom of absorber column **16**.

[0083] Overhead distillation stream **46** is withdrawn from contacting device absorber column **16** at -64°F [-53°C] and flows to reflux condenser **17** where it is cooled to -78°F [-61°C] and partially condensed by heat exchange with the cold LNG (stream **41a**) as described previously. The partially condensed stream **46a** enters reflux separator **18** wherein the condensed liquid (stream **49**) is separated from the uncondensed vapor (stream **48**). The liquid stream **49** from reflux separator **18** is pumped by reflux pump **19**

to a pressure slightly above the operating pressure of absorber column **16** and stream **49a** is then supplied as cold top column feed (reflux) to absorber column **16**. This cold liquid reflux absorbs and condenses the C_3 components and heavier hydrocarbon components from the vapors rising in absorber column **16**.

[0084] The residue gas (stream **48**) leaves reflux separator **18** at -78°F [-61°C], is heated to -40°F [-40°C] in cross exchanger **29** (stream **48a**), and is compressed by compressor **28** to sales line pressure (stream **48b**). Following cooling to -37°F [-38°C] in cross exchanger **29**, stream **48c** is heated to 30°F [-1°C] using low level utility heat in heat exchanger **30** and the residue gas product (stream **48d**) flows to the sales gas pipeline at 1315 psia [9,067 kPa(a)] for subsequent distribution.

[0085] A summary of stream flow rates and energy consumption for the process illustrated in FIG. 13 is set forth in the following table:

Table XIII

(FIG. 13)

Stream Flow Summary - Lb. Moles/Hr [kg moles/Hr]

Stream	Methane	Ethane	Propane	Butanes+	Total
41	9,524	977	322	109	10,979
44	850	534	545	127	2,058
45	850	528	239	18	1,637
46	28,574	3,952	83	0	32,732
49	19,050	2,981	67	0	22,174
48	9,524	971	16	0	10,558
47	0	6	306	109	421

Recoveries*			
Propane	95.05%		
Butanes+	99.98%		
Power			
LNG Feed Pump	616	HP	[1,013 kW]
Reflux Pump	103	HP	[169 kW]
Overhead Pump	74	HP	[122 kW]
Residue Gas Compressor	1,424	HP	[2,341 kW]
Totals	2,217	HP	[3,645 kW]
Low Level Utility Heat			
LNG Heater	32,453	MBTU/Hr	[20,965 kW]
Residue Gas Heater	12,535	MBTU/Hr	[8,098 kW]
Totals	44,988	MBTU/Hr	[29,063 kW]
High Level Utility Heat			
Deethanizer Reboiler	8,218	MBTU/Hr	[5,309 kW]

* (Based on un-rounded flow rates)

[0086] Comparing Table XIII above for the FIG. 13 embodiment of the present invention with Table XII for the FIG. 12 embodiment of the present invention shows that the liquids recovery is the same for the FIG. 13 embodiment. Since the FIG. 13 embodiment uses a pump (overhead pump **25** in FIG. 13) rather than a compressor (overhead compressor **23** in FIG. 12) to route the overhead vapor from fractionation stripper column **21** to contacting device absorber column **16**, less power is required by

the FIG. 13 embodiment. However, since the resulting stream **45d** supplied to absorber column **16** is not fully vaporized, more liquid leaves absorber column **16** in bottoms stream **44** and must be stripped of its methane and C_2 components in stripper column **21**, increasing the load on reboiler **22** and increasing the amount of high level utility heat required by the FIG. 13 embodiment of the present invention compared to the FIG. 12 embodiment. The choice of which embodiment to use for a particular application will generally be dictated by the relative costs of power versus high level utility heat and the relative capital costs of pumps and heat exchangers versus compressors.

Other Embodiments

[0087] In the FIG. 13 embodiment of the present invention, the partially heated LNG leaving reflux condenser **17** (stream **41b**) supplies the final cooling to the overhead vapor (stream **45a**) from fractionation stripper column **21**. In some instances, there may not be sufficient cooling available in stream **41b** to totally condense the overhead vapor. In this circumstance, an alternative embodiment of the present invention such as that shown in FIG. 14 could be employed. Heated liquefied natural gas stream **41e** is directed into contacting device absorber column **16** wherein distillation stream **46** and liquid stream **44** are formed and separated. Liquid stream **44** is directed into fractionation stripper column **21** wherein the stream is separated into vapor stream **45** and liquid product stream **47**. Vapor stream **45** is cooled sufficiently to partially condense it in cross exchanger **24** and heat exchanger **13**. An overhead separator **26** can be used to separate the partially condensed overhead stream **45b** into its respective vapor fraction (stream **50**) and liquid fraction (stream **51**). Liquid stream **51** enters overhead pump **25** and is

pumped through cross exchanger **24** to heat it and partially vaporize it (stream **51b**). Vapor stream **50** is compressed by overhead compressor **23** (with optional heating before and/or cooling after compression via heat exchangers **31** and/or **32**) to raise its pressure so that it can be combined with the outlet from cross exchanger **24** to form combined stream **45c** that is thereafter supplied to absorber column **16** at a lower column feed point. Alternatively, as shown by the dashed line, some or all of the compressed vapor (stream **50c**) may be supplied separately to absorber column **16** at a second lower column feed point. Some applications may favor heating the vapor prior to compression (as shown by dashed heat exchanger **31**) to allow less expensive metallurgy in compressor **23** or for other reasons. Cooling the outlet from overhead compressor **23** (stream **50b**), such as in dashed heat exchanger **32**, may also be favored under some circumstances.

[0088] Some circumstances may favor cooling the high pressure stream leaving overhead compressor **23**, such as with dashed heat exchanger **24** in FIG. 15. It may also be desirable to heat the overhead vapor before it enters the compressor (to allow less expensive metallurgy in the compressor, for instance), such as with dashed cross exchanger **24** in FIG. 16. The choice of whether to heat the inlet to the overhead compressor and/or cool the outlet from the overhead compressor will depend on the composition of the LNG, the desired liquid recovery level, the operating pressures of absorber column **16** and stripper column **21** and the resulting process temperatures, and other factors.

[0089] Some circumstances may favor using a split feed configuration for the LNG feed (as disclosed previously in FIGS. 10 and 11) when using the two column

embodiments of the present invention. As shown in FIGS. 15 through 18, the partially heated LNG (stream **41b** in FIGS. 15 and 16 and stream **41c** in FIGS. 17 and 18) can be divided into two portions, streams **42** and **43**, with the first portion in stream **42** supplied to contacting device absorber column **16** at an upper mid-column feed point without any further heating. After further heating, the second portion in stream **43** can then be supplied to absorber column **16** at a lower mid-column feed point, so that the cold liquids present in the first portion can provide partial rectification of the vapors in the second portion. The choice of whether to use the split feed configuration for the two column embodiments of the present invention will generally depend on the composition of the LNG and the desired liquid recovery level.

[0090] In the FIG. 17 embodiment using a split feed configuration for the LNG feed, liquid stream **44** is directed into fractionation stripper column **21** wherein the stream is separated into vapor stream **45** and liquid product stream **47**. The vapor stream is cooled in cross exchanger **24** and heat exchanger **33** to substantial condensation. The substantially condensed stream **45b** is pumped to higher pressure by pump **25**, heated in cross exchanger **24** to vaporize at least a portion of it, and thereafter supplied as stream **45d** to contacting device absorber column **16** at a lower column feed point.

[0091] In the FIG. 18 embodiment using a split feed configuration for the LNG feed, vapor stream **45** is cooled in cross exchanger **24** and heat exchanger **33** sufficiently to partially condense it and is thereafter separated in overhead separator **26** into its respective vapor fraction (stream **50**) and liquid fraction (stream **51**). Liquid stream **51** enters overhead pump **25** and is pumped through cross exchanger **24** to heat it and

partially vaporize it (stream **51b**). Vapor stream **50** is compressed by overhead compressor **23** (with optional heating before and/or cooling after compression via heat exchangers **31** and/or **32**) to raise its pressure so that it can be combined with the outlet from cross exchanger **24** to form combined stream **45c** that is thereafter supplied to absorber column **16** at a lower column feed point. Alternatively, as shown by the dashed line, some or all of the compressed vapor (stream **50c**) may be supplied separately to absorber column **16** at a second lower column feed point. Some applications may favor heating the vapor prior to compression (as shown by dashed heat exchanger **31**) to allow less expensive metallurgy in overhead compressor **23** or for other reasons. Cooling the outlet from overhead compressor **23** (stream **50b**), such as in dashed heat exchanger **32**, may also be favored under some circumstances.

[0092] Reflux condenser **17** may be located inside the tower above the rectification section of fractionation tower **16** or absorber column **16** as shown in FIG. 19. This eliminates the need for reflux separator **18** and reflux pump **19** shown in FIGS. 10 through 18 because the distillation stream is then both cooled and separated in the tower above the fractionation stages of the column. Alternatively, use of a dephlegmator (such as dephlegmator **27** in FIG. 20) in place of reflux condenser **17** in FIGS. 10 through 18 eliminates the need for reflux separator **18** and reflux pump **19** and also provides concurrent fractionation stages to supplement those in the upper section of the column. If the dephlegmator is positioned in a plant at grade level, it can be connected to a vapor/liquid separator and the liquid collected in the separator pumped to the top of the distillation column (either fractionation tower **16** or contacting device absorber column

16). The decision as to whether to include the reflux condenser inside the column or to use a dephlegmator usually depends on plant size and heat exchanger surface requirements.

[0093] It also should be noted that valves **12** and/or **15** could be replaced with expansion engines (turboexpanders) whereby work could be extracted from the pressure reduction of stream **42** in FIGS. 10, 11, and 15 through 18**,** stream **43b** in FIGS. 10, 11, and 15 through 18, and/or stream **41d** in FIGS. 12 through 14. In this case, the LNG (stream **41**) must be pumped to a higher pressure so that work extraction is feasible. This work could be used to provide power for pumping the LNG stream, for compression of the residue gas or the stripper column overhead vapor, or to generate electricity. The choice between use of valves or expansion engines will depend on the particular circumstances of each LNG processing project.

[0094] In FIGS. 10-20, individual heat exchangers have been shown for most services. However, it is possible to combine two or more heat exchange services into a common heat exchanger, such as combining heat exchangers **13**, **14**, and **24** in FIG. 14 into a common heat exchanger. In some cases, circumstances may favor splitting a heat exchange service into multiple exchangers. The decision as to whether to combine heat exchange services or to use more than one heat exchanger for the indicated service will depend on a number of factors including, but not limited to, LNG flow rate, heat exchanger size, stream temperatures, etc.

[0095] It will be recognized that the relative amount of feed found in each branch of the split LNG feed to fractionation tower **16** or absorber column **16** will depend on

several factors, including LNG composition, the amount of heat which can economically be extracted from the feed, residue gas delivery pressure, and the quantity of horsepower available. More feed to the top of the column may increase recovery while increasing the duty in reboiler **22** and thereby increasing the high level utility heat requirements. Increasing feed lower in the column reduces the high level utility heat consumption but may also reduce product recovery. The relative locations of the mid-column feeds may vary depending on LNG composition or other factors such as the desired recovery level and the amount of vapor formed during heating of the feed streams. Moreover, two or more of the feed streams, or portions thereof, may be combined depending on the relative temperatures and quantities of individual streams, and the combined stream then fed to a mid-column feed position.

[0096] While there have been described what are believed to be preferred embodiments of the invention, those skilled in the art will recognize that other and further modifications may be made thereto, e.g. to adapt the invention to various conditions, types of feed, or other requirements without departing from the spirit of the present invention as defined by the following claims.